

Hydrogen-Enriched Syngas from Biomass Steam Gasification for Use in Land-Based Gas Turbine Engines

John Dascomb and Anjaneyulu Krothapalli

Abstract This paper presents the results from an experimental study on the production of hydrogen-enriched syngas through un-catalyzed steam biomass gasification. Wood pellets were gasified using a 100 kW_{th} fluidized bed gasifier at temperatures up to 850 °C. The syngas hydrogen concentration was found to increase with both bed temperature and steam-to-biomass weight ratio, reaching a maximum of 51 %. The overall energy conversion to syngas (based on heating value) also increased with bed temperature but was inversely proportional to the steam-to-biomass ratio. The maximum heating value measured was 16.1 MJ/m³, occurring at a temperature of 697 °C, a S/B of 0.7, and a residence time of 1.9 s. The maximum energy conversion to syngas was found to be 68 %. Further analysis revealed that to produce syngas with high hydrogen concentration, inert-fluidized gasification was the most efficient method tested in this study. While fluidized bed gasification with CaO was found to be less efficient than inert-bed gasification, it does produce syngas with higher hydrogen concentration.

Keywords Biomass steam gasification • Syngas • Gas turbine engines • Energy conversion

1 Introduction

Worldwide, the use of energy from all sources will increase over the next few decades, primarily because of the demand driven by high GDP growth nations, such as China and India. Given the expectation that crude oil prices will remain relatively high, and petroleum and other liquid fuels are the world's slowest growing energy sources. Additionally, the concern about environmental consequences of green

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house gas emissions leads to a number of national governments to provide incentives in support of the development of alternative energy sources, making renewable sources of energy as the fastest growing in the outlook.

Gas turbines have emerged as the best means of transforming heat into mechanical energy and are now key components of the most efficient electrical generating systems. Hydraulic fracturing, or fracking, has increased the shale gas production, which accelerated the use of gas turbines for power production in lieu of coal power. However, many countries such as India and China do not have access to the shale gas and as a result will depend on alternative sources such as the syngas produced from either coal or biomass gasification. The more environmentally accepted solution would be syngas that is produced from biomass gasification. To be considered interchangeable with conventional fossil fuels (natural gas) and to ensure maximum flexibility, syngas heating value needs to be above 11 MJ/m^3 . A high hydrocarbon content corresponds to a higher heating value for the syngas. Hence, a hydrogen-enriched synthesis gas with a higher heating value will help to mitigate the use of natural gas in land-based gas turbine applications. Hence, the subject of this paper is to present a cost effective means of biomass-derived synthesis gas production that is suitable for use in land-based gas turbines for power production.

The preferred feedstock for the current purpose is the cellulosic biomass that includes plant matters such as wood chips, switchgrass, wheat straw, corn stovers, sorghum, oilseed crop meals, and agriculture wastes. Cellulosic biomass is an advantageous feedstock because it is not an edible food crop feedstock. The cellulosic biomass consists primarily of cellulose, hemicellulose, and lignin, and these compounds pose significant challenges to conventional chemical and microbial processing methods. The process described in this paper can utilize different types of cellulosic biomass including lignocellulosic biomass and will ensure a continuous and renewable feedstock supply that is affordable. Because of the ability to convert cellulosic biomass, the proposed process has a significant advantage in full plant utilization. Other technologies have not addressed this important consideration.

1.1 Dual Fluidized Bed (DFB) Steam Gasification

Gasification is a well-proven technology that has been employed in various forms for almost 200 years (Klass 1998). It is essentially an oxygen limited thermochemical conversion of carbonaceous material to a useable gaseous fuel, synthesis gas or “syngas” consisting primarily of hydrogen (H_2) and carbon monoxide (CO), with lesser amounts of carbon dioxide (CO_2), methane (CH_4), higher hydrocarbons (C_2+), water (H_2O), and nitrogen (N_2). The oxidant used can be air, pure oxygen, or steam. Air-based gasifiers typically produce a product gas containing relatively high concentration of nitrogen with a low heating value (LHV) between 3 and 5 kJ/kg ($4\text{--}7 \text{ MJ/Nm}^3$)². Steam-based gasifiers, on the other hand, produced a product gas containing relatively high concentration of hydrogen and CO with heating values

between 10 and 20 kJ/kg (10–20 MJ/Nm³) (Paisley et al. 2001). Such a gas can be used in variety of applications, such as turbine-based power generation units, fuels, chemicals, and hydrogen production, with relatively simple downstream conversion technologies. A note that a high calorific value gas such as natural gas has a LHV of 30 MJ/Nm³.

In a standard directly heated gasification process (Nguyen et al. 2010), the pyrolysis and gasification reactions occur in a single vessel. An oxidant, air, or oxygen combusts a portion of the biomass to provide the heat required for the endothermic reactions. Pyrolysis requires about 15 % of the heat of combustion of the feed to raise the reaction temperature and vaporize the products. In these systems, the reactor temperature is controlled by the oxidant feed rate. If the air is used as the oxidant, the product gas has a low heating value of 4–5 MJ/m³ due to nitrogen dilution as shown in Table 1 (Roesch et al. 2011; Roesch 2011).

The results shown in the Table 1 were obtained in our laboratory, using an Ankur Scientific downdraft gasifier (Roesch 2011). As the gas composition results, shown in Table 1, indicate that a large percentage (~50 %) of the syngas contained high concentrations of N₂ for all the feedstocks considered in this study. The relatively low percentages of H₂, CO, and CH₄ are mainly responsible for the low lower heating values of the syngas.

By gasifying biomass with an N₂-free agent, the product gas will not be diluted with large concentrations of the inert gas. Using either O₂ or H₂O as the gasifying agent produces medium grade product gas (~11 MJ/Nm³) (Ciferno and Marano 2002). In addition, catalysts in the gasifier and downstream can aid in the conversion of tars and char to usable gas. With these and other advanced techniques, a gas with a LHV of up to 20 MJ/Nm³ is possible. The drawback to this higher quality gas is the absence of the heat produced by the burning of char. The reactions that break up the higher carbon chains in the biomass to produce H₂ and CO are highly endothermic. Heat must be provided to the process without polluting the produced gas with combustion products.

The best solution found to the problem of product stream dilution has been to burn the char produced in the reactor in a separate chamber and transfer the heat

Table 1 Syngas composition for different feedstocks

Pelletized feedstock	Component mol% dry basis						
	H ₂	O ₂	N ₂	CH ₄	CO	CO ₂	HHC
Alfalfa	12.823	0.020	53.004	2.067	12.245	18.589	1.253
Algae	10.768	0.035	59.395	1.564	9.911	17.317	1.011
Field grass	12.454	0.023	51.621	2.860	13.630	17.795	1.617
Hemp	10.362	0.019	58.546	2.098	12.622	14.891	1.463
Miscanthus	9.147	0.039	53.152	2.685	17.509	16.194	1.274
Peanut shells	13.410	0.033	51.017	2.635	15.833	15.735	1.337
Pine	12.300	0.498	53.422	2.696	16.568	13.368	1.150
Municipal waste	11.323	0.029	59.493	1.899	9.615	16.257	1.384

back to the gasification process, commonly referred to as indirectly heated gasification technology. A successful scheme to achieve the indirectly heated gasification process is the dual fluidized bed (DFB) system as shown in Fig. 1. The unreacted char and bed material in the gasifier is carried to a combustion chamber where air is injected. The heat produced is then transferred back into the gasifier by the bed material. The gasifier utilizes a fluidized bed usually consisting of sand. When a gas is passed through the bed of fine solids, the bed takes on the characteristics of a liquid, allowing it to flow between the respective chambers. The continuously circulating bed greatly increases heat transfer and interaction time between the biomass and gasifying agent, while the sand breaks up the biomass quicker leading to higher gas production. The bed material easily flows when aerated, and the liquidity of the bed aids in the transfer of char to the different chambers of the system. This approach separates the combustion reaction from the remaining gasification reactions producing a product gas that has insignificant percentage of N_2 and has a heating value within a range of 15–18 MJ/Nm³.

A list of dual fluidized bed gasification facilities that are used both in research and power plant scenarios are listed in Table 2 (Paisley et al. 2001; Kappotz et al. 2009, 2011; van der Meijden et al. 2007a, b; Corella et al. 2007; Herguido et al. 1992; Xu et al. 2006; Wei et al. 2006; Matsuoka et al. 2008). Most comprehensive studies of DFB gasification are more recently carried out at Vienna University of Technology in Austria (Koppotz et al. 2009, 2011) using both laboratory and pilot-scale facilities. The key findings from their work using the CO_2 absorption enhanced reforming process were a significantly higher H_2 content (50–60 %) in the producer gas as compared to the conventional dual fluidized bed steam gasification. The high hydrogen concentrations were obtained at gasification temperatures of about 700 °C and using repeated cycles of carbonation and calcination.

All the systems cited above use steam as the gasification agent in a typical DFB system. Because of the high heat of vaporization of steam and gasification being

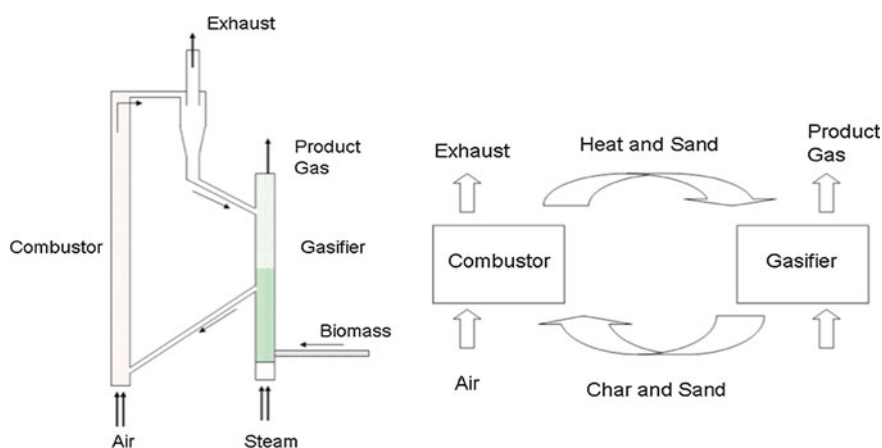


Fig. 1 Flow schematics for a dual fluidized bed gasification system

Table 2 Dual fluidized bed gasification facilities

Gasification facility	Size	Bed material	H ₂ (vol%)	CO (vol%)	CH ₄ (vol%)	HV MJ/m ³
McNeil plant, Vermont; DFB	50 MW		22	44.4	15.6	17.3 (HHV)
Gussing, Austria; demo DFB plant	8 MW	Limestone	50.6	16.5	12.9	
ECN (DEN) pilot-scale DFB	800 kW	Austrian olivine	21.2	37.2	12.1	
FSU, Florida; single bed	100 kW	Silica	28.7	43.5	14.6	16.5 (HHV)
University of Vienna; lab-scale DFB	100 kW	Calcite	65.1	9.3	8.8	13.1–16.5 (LHV)
University of Zaragoza, DFB	45 kW	Catalyst	30.1			15.8 (LHV)
ECN (DEN) lab-scale DFB	30 kW	Austrian olivine	27.3	27.5	9.5	
University of Zaragoza, single bed	20 kW	Silica	50	22	6	12.75 (LHV)
Yokohoma, Japan; DFB	25 kW	Silica	23.2	37.8	16.9	15.8 (HHV)
Dalin, China; dual moving beds	5 kW	Olivine	40	30	9	
Onogawa, Japan; circulating DBFB	0.22 kW	Alumina	35	27.5	7.5	

endothermic, most of the systems cited above require additional heating to maintain required reactor bed temperature. Since steam generation requires a significant fraction of the energy and the affordability of the process is questionable. Hence, only the 8-MW Gussing, Austria plant is currently in operation for power production (Koppatz et al. 2011). As a result, a careful investigation of process thermodynamics along with energy balance needs to be carried out for the design of an optimal DFB gasifier. With this in mind, a program of research was undertaken at FSU to produce affordable hydrogen-enriched synthesis gas using multiple biomass feedstock.

Figure 2 shows a sustainable DFB gasification facility schematic for a complete hydrogen-enriched syngas/H₂ production plant. The process begins with renewable biomass feedstock. The advantage of this facility is that it can process any and all parts of a biomass, even cellulosic wastes. This increases the total energy efficiency of the syngas production process. The biomass is chipped and dried with the combustion exhaust before being inserted into the gasification reactor. Inside the reactor, superheated steam is injected at up to 800 °C to fluidize and react with the bed material. Char formed in the gasification reaction is transferred to the combustor where it is burned by the addition of air. The heat generated by this combustion is then put back into the gasifier to continue the reactions. The superheated steam and product gas mixture leaves the gasification reactor where it is cooled in a heat exchanger exiting at a temperature of 350–400 °C. The gas then goes through the

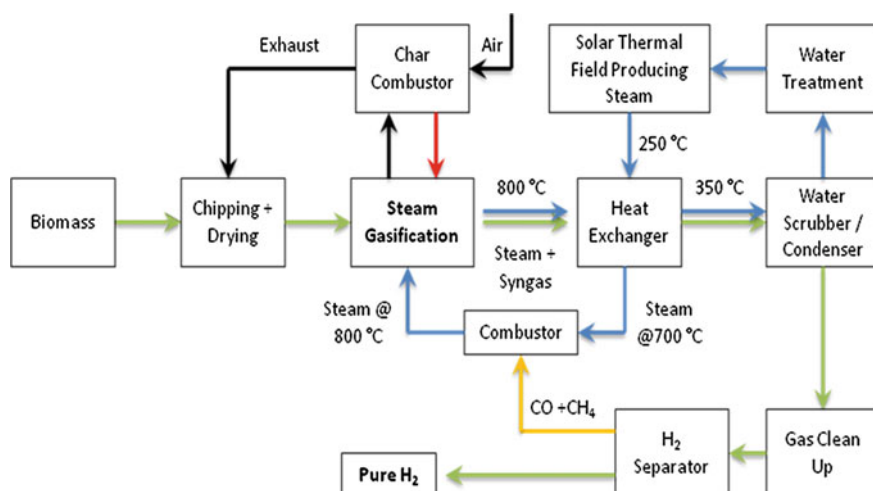


Fig. 2 Sustainable hydrogen-enriched syngas/hydrogen production process

scrubber/condenser that eliminates any tars and condenses out the fluidizing agent—the steam. The condensed water is processed before being reheated in a solar thermal field and the heat exchanger. The heated steam is brought up to its final temperature in a combustor before being once again injected into the gasifier. Back at the water scrubber, the cooled product (synthesis) gas is cleaned some more before the H_2 is separated out. All other combustibles are reused in the facility to superheat steam or heat the gasification reactor.

2 Experimental Facility and Procedures

A 100 kW-capacity dual fluidized bed thermal steam gasifier is proposed at FSU. As the first step, a single reactor bed was built to test many of the sub-systems of the facility and determine the thermal efficiency of the chemical process. A schematic of the single bed system is shown in Fig. 3. In the absence of the char combustor, heat is externally supplied to the reactor using series of electric heaters controlled with variable transformers. To precisely control the amount of steam used in the reactor, a standard electric boiler/superheater unit was installed. The advantage to both these changes is it gives the operator a highly controlled and measured process. The hot synthesis gas is then sent through a cyclone filter to remove any solid particles, followed by a heat exchanger, where the gas is cooled to about 400 °C. The gas is then passed through a venturi water scrubber/condenser and filters to further improve the gas quality. The resulting gas chemical characteristics are monitored in situ using a gas chromatograph before it is burned in a controlled Bunsen burner flame. The complete thermal characteristics of the system, as shown in Fig. 4, will be monitored and controlled.

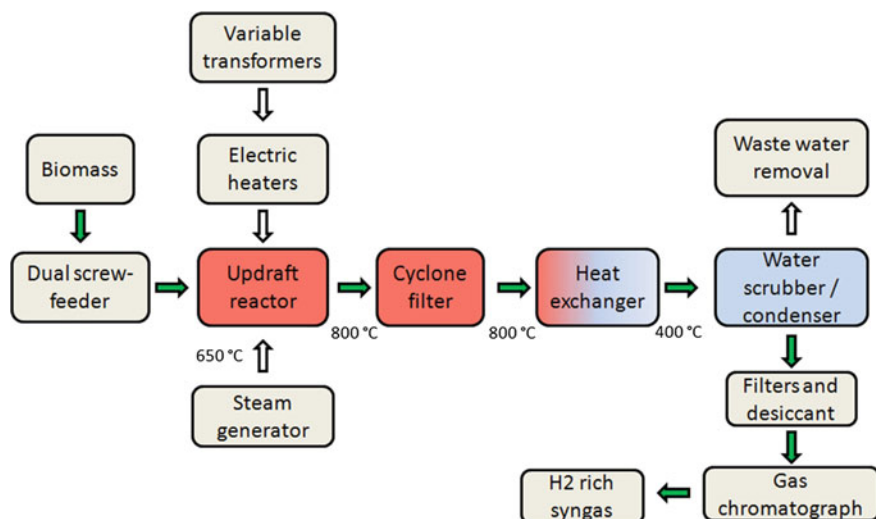


Fig. 3 Single bed hydrogen-enriched synthesis gas production process

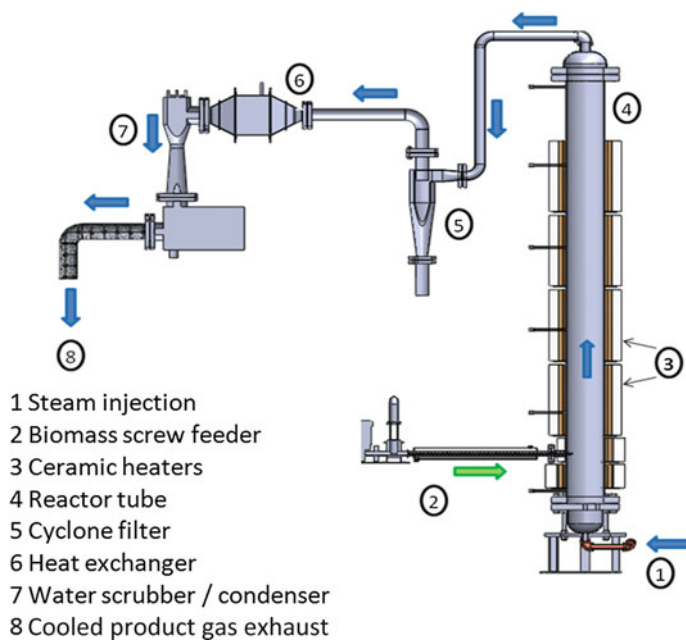


Fig. 4 Single bed hydrogen-enriched synthesis gas production plant

The main control parameters that govern the biomass conversion chemical process are as follows: bed temperature, steam-to-biomass ratio, biomass mass throughput, and catalytic properties. The energy balance of the system is determined from the measurement of the all the energy inputs such as the heating value of the biomass, steam energy, electric energy, and the heating value of the synthesis gas. Using measurement and analysis tools, the objective is to determine the most cost-efficient reactor conditions for desired gas composition and the heating value.

The settling chamber and reactor are heated using independently controlled radiant heaters. The heating system has a maximum capacity of 29 kWe and provides all the necessary energy for maintaining bed temperature and gasifying the biomass feedstock. The system is capable of gasifying up to 20 kg/h (115 kWth) of biomass pellets at 650 °C and 9 kg/h (50 kWth) at 850 °C.

Steam enters the reactor through 5 tuyeres (gas injectors). The tuyeres were constructed out of steel piping (24 mm OD) and protrude 76 mm into the gasifier bed. Through holes (1 mm) were drilled radially into the tuyeres to ensure uniform steam injection while not allowing material to flow back into the settling chamber. The gasifier reactor is filled with inert silica sand to a static height of 1.0 m. The pure SiO₂ sand has an average particle size of 0.28 mm. When fluidized, the bed height reaches 1.5–2.5 m depending on the steam flow rate. The reactor was constructed from 316 stainless steel with an inner diameter of 0.21 m and a height of 3.1 m.

The biomass used for gasification is stored in a sealed hopper before being injected into the reactor. The pellets have an average diameter of 8 mm and maximum length of 32 mm. The pellets are carried from the hopper by a metering auger. The pellets are then fed by a fast injection feeder into the reactor just above the steam injection point. A water-cooled jacket prevents premature pyrolysis in the feeding system. After the biomass and steam mixture pass through the reactor, a cyclone mechanical filter removes any solid bed materials and unreacted char from the gas stream. The gas then flows to a condenser where it is cooled to 30 °C. In addition to condensing out steam, many of the heavy tars and ash remaining in the syngas are also removed. Following the condenser, the product gas composition is analyzed using a gas chromatograph. A micro-pump carries a small amount of the gas to a filter system where it is cleaned and dried before analysis. The bulk syngas stream is then filtered by a bed of sawdust and fabric filter. The volumetric flow rate of the filtered stream is measured using a custom venturi flow meter before being disposed.

For each experiment, the reactor is first heated to the desired testing temperature by the radiant heaters. The heaters are connected to variable transformers, which are controlled by LabView process software. All facility control and measurement hardware utilized a NI compact DAQ chassis and input/output modules. The calibrated transformers tune the voltage sent to each of the heaters to maintain uniform temperature along the length of the reactor. To achieve this, seven K-type thermocouples are evenly spaced along the reactor. The temperatures are recorded using a 24-bit thermocouple module (NI-9213). Once the reactor becomes thermally stable, the electrical load of the heaters is measured to determine the thermal

Table 3 Biomass feedstock Analysis

Parameter	Value	Units
C	51.34	Moist. free wt%
O	42.50	Moist. free wt%
H	5.98	Moist. free wt%
N	0.17	Moist. free wt%
S	<0.01	Moist. free wt%
Ash	<0.6	Moist. free wt%
Moisture	5.82	wt%
LHV	20,242	kJ/kg

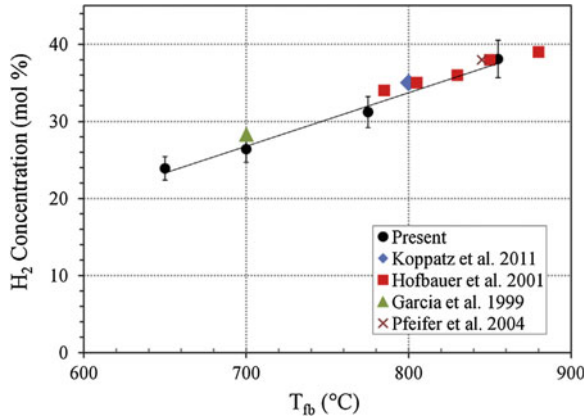
losses. The calibrated variable transformers have a relative error of 2.5 %. The current draw of the heaters is measured by ammeters that have a relative error of 5.0 %. Next, the steam generator and superheaters are started. The steam flow rate is determined by measuring the liquid feed rate (relative error of 4.3 %) which intermittently feeds the boiler. The flow rates used for calculations are averaged over an hour of testing. The electrical load of the boiler and superheaters is not directly measured, but calculated using the enthalpy gain from liquid water at standard temperature and pressure to its state entering the settling chamber.

The calibrated metering feeder controls the biomass feed rate. The feeder has a measured error of 6.9 %, which is caused by variations in pellet size, hopper fill height, and pellet structural integrity. The pellets contain no binders and tend to break up when agitated. The biomass feedstock used for all testing is north Florida pine. Table 3 shows the chemical analysis for the feedstock. Green Circle Bio Energy manufactures the pellets locally from native trees. The pellets have a low ash content of less than 0.6 %, which helps prevent conglomeration in the reactor. After gasification and cooling, the syngas mixture is analyzed by an Agilent Micro GC 3000, which has a relative error of 6.4 %. The chromatograph detects the dry relative concentrations of all compounds with atomic mass less than propylene (C_3H_6). During operation, samples are drawn from the produced stream every 4 min for analysis. Once the syngas is filtered and analyzed, the volume flow rate is then measured using a custom venturi type flow meter that has a relative error of 5.3 %. Another K-type thermocouple measures the gas density at the venturi flow meter.

3 Results and Discussion

To verify gasifier results, the syngas hydrogen concentration and cold gas efficiency are compared to published data. Figure 5 compares the hydrogen concentration found in this study to published results at similar steam-to-biomass (S/B) ratios and residence times. The agreement in the data confirms gasifier operation and the gas analysis techniques used in this study.

Fig. 5 Comparison of hydrogen concentration at similar run conditions ($S/B = 0.8 \pm 0.25$, $\tau = 2.0 \pm 0.5$)

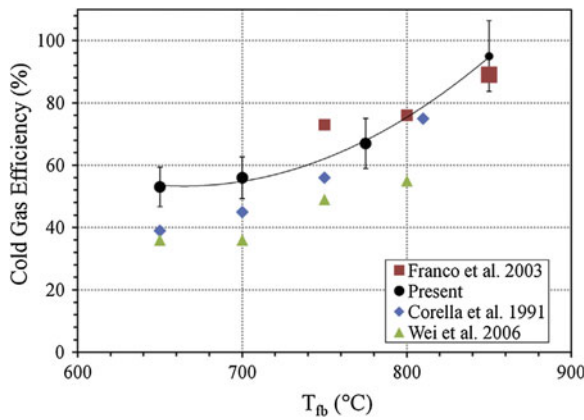


The cold gas efficiency is also compared to published data at similar temperatures and S/B ratios and is shown in Fig. 6 (Corella et al. 1991; Franco et al. 2003; Xu et al. 2004). The cold gas efficiency is reported less frequently than hydrogen concentration, as a result the residence times could not be perfectly matched. The cold gas efficiency is defined as follows:

$$\text{Cold Gas Efficiency (CGE)} = \frac{\dot{m}_{\text{syngas}} \cdot \text{HHV}_{\text{syngas}}}{\dot{m}_b \cdot \text{HHV}_b}$$

where subscript b denotes the biomass. Variation in residence time, gasifier shape, feedstock, feedstock entry points, and method of heating all effect CGE. Wei et al. (2006) used a dual fluidized bed gasifier system that burned a portion of the char in a separate chamber. This is a possible reason why their reported efficiency is lower. Additionally, Corella et al. (1991) injected biomass at the top of the fluidized bed, resulting in lowered gasification efficiency. Despite these differences, the trend of

Fig. 6 Comparison of temperature effect on cold gas efficiency ($S/B = 0.8$, $\tau = 2.2$ s)



increased CGE with increased temperature holds for all of the facilities. The CGE found in the present study matches (within limits of error) the results obtained by Franco et al. at temperature above 750 °C. The agreement in the CGE values confirms proper operation of the gasifier energy measurement systems.

3.1 Syngas Composition

The effect of the reactor temperature on syngas composition was studied between 650 and 850 °C. Figure 7 shows the concentration of the major syngas constituents for moderate S/B ratios and residence times. At 650 °C, carbon monoxide was the most dominant syngas component at 32 mol%, followed closely by hydrogen at 30 mol%. When the reactor temperature was raised, the hydrogen concentration increased, while all of the other constituents gradually decreased. At 775 °C, hydrogen was the most dominant species at 38 mol% and carbon monoxide levels had dropped to 30 mol%. This trend continued when the reactor temperature was raised to 850 °C, where a hydrogen concentration of 44 mol% was recorded. Part of the reason for the increase in hydrogen concentration was the cracking of hydrocarbons. Methane levels dropped from 13 to 10 mol%, and higher hydrocarbon (HHC) levels dropped from 5 to 3 mol%.

The effect of temperature on HHC concentration is shown in Fig. 8. Ethylene (C_2H_4) levels increased as the reactor temperature was raised from 650 to 775 °C before falling to its lowest level at 850 °C. The increase in ethylene concentration could have been from the cracking of propylene (C_3H_6). Propylene contains two carbon atoms joined by a double bond (like ethylene) and with a methyl (methane) group attached to it. All of the other HHCs decreased in concentration with each subsequent increase in temperature. The resulting sum of the HHC's decreased with

Fig. 7 Effect of temperature on gas composition (S/B = 2.0 ± 0.1 , $\tau = 2.0 \pm 0.4$ s)

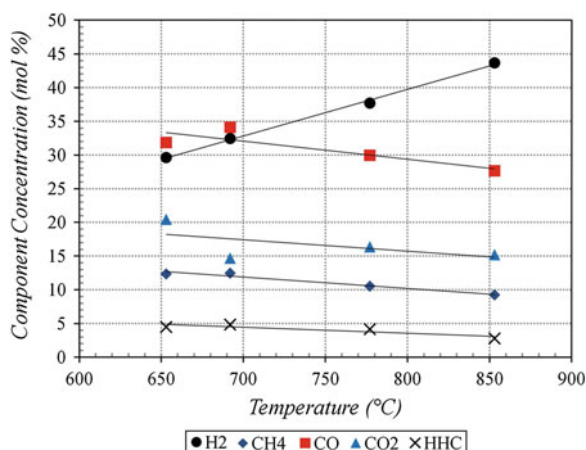
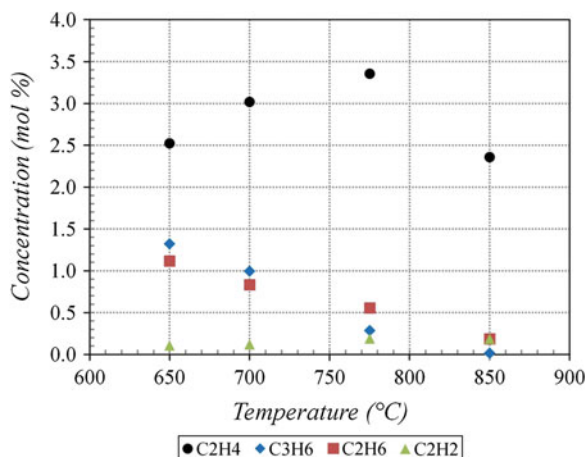


Fig. 8 Effect of temperature on higher hydrocarbon concentrations
($S/B = 2.0 \pm 0.1$,
 $\tau = 2.0 \pm 0.4$ s)



temperature. The hydrocarbons crack at high temperatures to produce lighter molecular weight compounds.

The trend of increasing hydrogen concentration with increased temperature held for all S/B ratios and residence times tested. Figure 9 shows the similar rate of increase in hydrogen concentration for three different bed conditions. Hydrogen concentration increased by an average of 7 mol% per 100 °C increase in reactor temperature. The hydrogen concentration increased due to the increase in gas-phase steady-state concentration as well as increased char reactivity. The char–steam gasification reaction is endothermic and occurs at very low rates at low temperatures. The results presented are in agreement with published data (Franco et al. 2003; Koppatz et al. 2011). A hydrogen concentration maximum was not found in the present experimental study, and an increase in temperature would almost certainly increase hydrogen concentration further. In order to achieve an average bed temperature of 900 °C, the reactor exterior temperature in the main heating

Fig. 9 Effect of temperature on hydrogen concentration for different S/B and τ

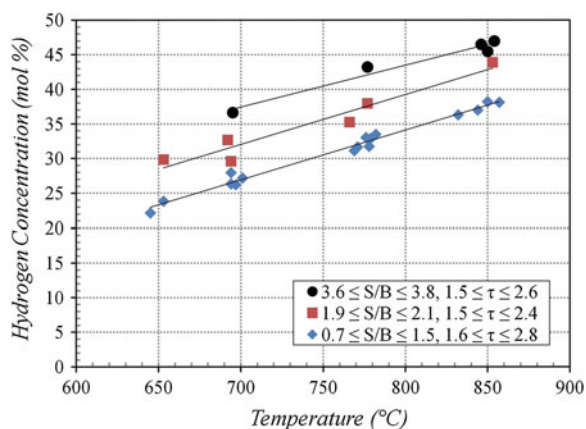
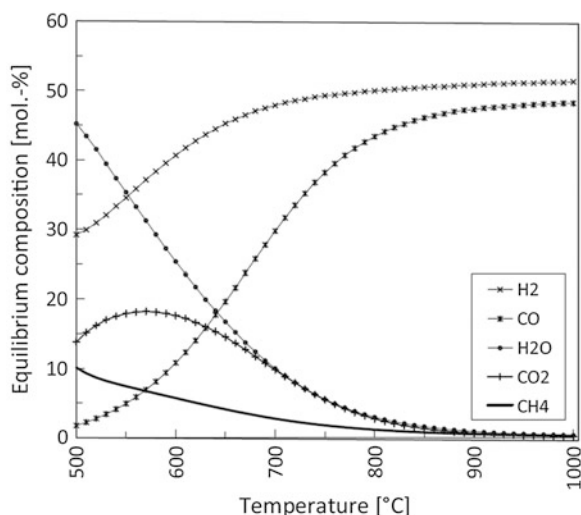


Fig. 10 Equilibrium syngas composition (Koppataz et al. 2011)



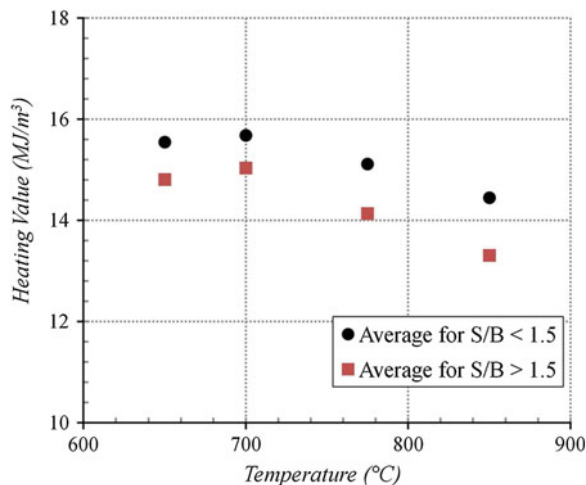
region was required to be in excess of 1,100 °C. At this temperature, corrosion of the reactor exterior surface occurred rapidly. As a result, the reactor surface would shed oxidized metal dust and flakes that accumulated in the gap between the heaters and reactor surface. This collection of solids created an electrical bridge between the heaters and the reactor, short circuiting the heaters. The resulting electrical discharge destroyed heaters and deformed the reactor surface. To test gasification at higher temperatures, a reactor material capable of withstanding such high temperatures is necessary, or oxygen must be eliminated from the environment.

Due to the difficulty of reaching gasification temperatures above 850 °C, most experimental facilities do not test above this temperature. Equilibrium models, such as the one by Koppataz et al. (2011) shown in Fig. 10, predict a slow increase in hydrogen concentration until 1,000 °C. Equilibrium models usually over predict hydrogen concentration at low temperatures due to slow reaction rates that resist changes in composition. When reactor temperature is increased to 800 °C and above, experimental results tend to more closely match equilibrium predictions. This explains the linear trend seen in the hydrogen concentration measured in this study. It is expected that at higher temperatures, the hydrogen concentration will taper off similar to the prediction of the equilibrium models.

3.2 Syngas Heating Value

The syngas heating value is an important parameter when determining the possible end uses for syngas. When used as a fuel to power an engine or turbine, the heating value must be within acceptable limits of the particular device. Hydrogen-enriched syngas can have a high heating value per unit mass, while having a low heating value

Fig. 11 Effect of temperature on syngas heating value



per unit volume compared to other gaseous fuels such as propane or natural gas. The relatively low density of syngas is due to the relatively high concentration of hydrogen, which has the lowest density of any gas. The heating value of gaseous fuels is usually reported on a volume basis. Hence, the following analysis is given as such. It is important to keep in mind that heating value presents qualitative information on the produced syngas and does not inform about the quantity of energy produced.

Figure 11 shows the effect of temperature on the syngas heating value. The heating value was highest at a reactor temperature of 700 °C. At this temperature, carbon monoxide was the most abundant gas constituent. The maximum heating value measured was 16.1 MJ/m³, occurring at a temperature of 697 °C, S/B of 0.7, and a residence time of 1.9 s. At this condition, the methane concentration was 13.4 mol%, which its highest measured, and the carbon monoxide concentration was also near its maximum at 38 mol%. The main reason that the heating was highest because of the hydrogen concentration being relatively low at 26.3 mol%. The heating value dropped as the reactor temperature was increased. The minimum heating value measured was 12.7 MJ/m³. This occurred at a reactor temperature of 853 °C, S/B of 2.9, and a gas residence time of 4.5 s. The hydrogen concentration during this particular experiment was the highest measured for un-catalyzed gasification at 51 mol%. The high hydrogen concentration increased the total combustible fraction and volume of the syngas, but decreased the heating value.

3.3 Energy Efficiency

One of the most important parameters determining the viability of steam gasification. is the thermodynamic conversion of energy, or energy efficiency (EE). This quantitative tool is a good first estimation of the cost of producing hydrogen-enriched

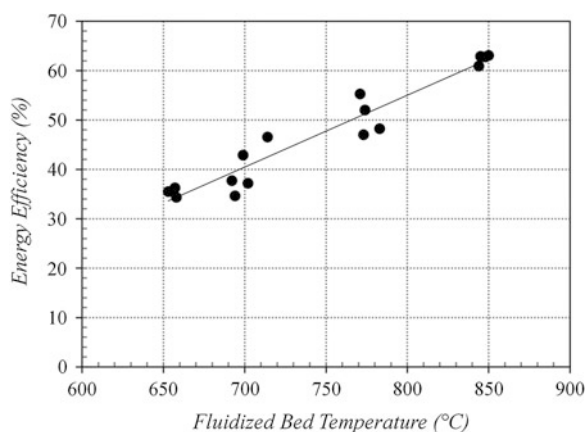
syngas. The energy efficiency describes the conversion of all energy sources to syngas (on a heating value basis). It is observed that the energy flow rate of the syngas increases with reactor temperature, gas residence time, and S/B ratio (Dascomb 2013). The drawback to this increased power production is that increasing any of the three major parameters increases the auxiliary heating requirement. Increasing reactor temperature obviously increased conduction losses, but it also increased the sensible heat of the gas. For facilities without a heat exchanger, such as the one used presently, that sensible energy is not recovered. The non-heat recovery energy efficiency gives a conservative estimate of the impact of increasing reactor temperature. To analyze the efficiency with a heat exchanger, one can simply assume a recovery efficiency of a heat exchanger.

It takes a considerable amount of energy to generate superheated steam. Steam has a high heat of vaporization, and most of the energy requirement is needed simply to make saturated steam. Because the reactor thermal losses were fixed at each temperature, when less biomass was injected into the reactor, the relative amount of thermal losses increased. In this way, increasing gas residence time increases the relative amount of thermal losses. In other words, the amount of joules lost per kg of biomass is greater when less kilograms of biomass are inserted into the reactor. The gasifier conditions that maximize the thermal conversion of all energy sources to syngas heating rate was discussed in a recent publication (Dascomb et al. 2013). The energy conversion efficiency used is defined as

$$\frac{\dot{m}_{\text{syngas}} \cdot \text{HHV}_{\text{syngas}}}{\dot{m}_b \cdot \text{HHV}_b + \dot{Q}_h + \dot{m}_b \cdot \Delta H_s}$$

Figure 12 shows the effect of temperature on energy efficiency. The energy conversion efficiency increased with increasing reactor temperature despite the increased conduction losses of the reactor and increased sensible heat requirement

Fig. 12 Effect of temperature on energy efficiency.
(S/B = 1.8 ± 0.4)



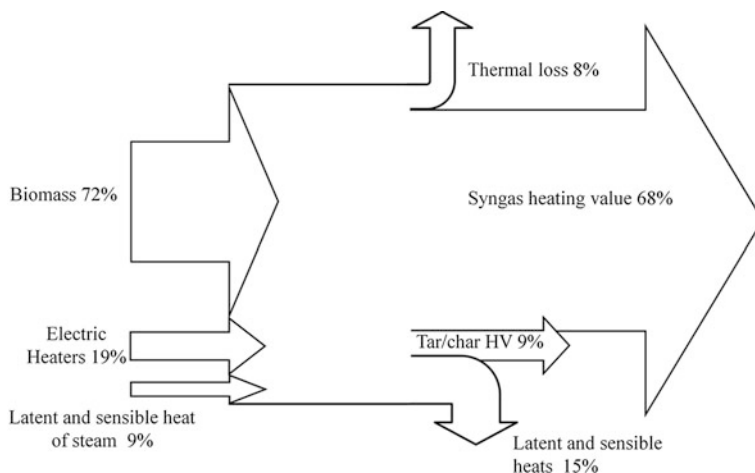


Fig. 13 Energy flow diagram for maximum energy efficiency case. ($T = 857\text{ }^{\circ}\text{C}$, $S/B = 0.75$, $\tau = 2.4\text{ s}$)

of the syngas. As shown in the previous section, the heating value of the syngas decreased as the reactor temperatures rose, but the total heating content (heating value \times flow rate) increased. An increase in temperature would most likely increase the conversion efficiency further. Adding a heat exchanger could also increase the energy efficiency. Recovery of any of the sensible heats of the steam or syngas mixture would reduce the required energy input and increase energy efficiency.

A heat flow diagram for the maximum energy efficiency case is shown in Fig. 13. The biomass heating value accounted for 72 % of the energy input, while 19 % was required to heat the reactor bed and 9 % for the generation of steam. The total process energy converted to chemical potential in the syngas was 68 %. Reactor thermal losses consumed 8 % of the input energy. This portion can be reduced by better insulating the reactor. The thermal loss during this test was 7.8 kW. Lower thermal losses were accomplished after this particular test with better insulation wrapping. The results presented here are using the measured losses to accurately portray testing.

The Latent and sensible heat contained by the steam and syngas accounted for 15 % of the total energy. This energy is available to be recycled back into the system using heat exchangers. Because of the higher tar content produced at low S/B ratios, some form of tar treatment will be needed to prevent reduction in heat exchanger efficiency. The easiest way to prevent tar buildup in a heat exchanger is only cool the syngas mixture down to $400\text{ }^{\circ}\text{C}$. In this way, most of the tars are prevented from condensing. While effective, this method will limit the amount of recoverable heat. Another option is to use a catalytic tar reformer. This heated vessel will remove most of the tars and thus allow the gas to pass through a heat exchanger that can utilize the full energy potential of the heated gas. Although this adds infrastructure and energy cost to maintain another bed at high gasification temperatures.

A carbon balance was performed to determine the amount of carbon not converted to syngas. At the maximum energy efficiency case, 18 % of the carbon in the biomass did not form usable syngas. This corresponded to any total energy loss of 9 %. There are multiple ways to increase the conversion of carbon and increase energy efficiency; higher gasification temperatures, the addition of catalysts, and a dual bed gasification system. The tar and char energy content is the least accurate value given in the energy balance due to the indirect form of measurement. It is necessary to note that the heat content of the outputs was 16 % larger than the inputs. This was most likely due to the accumulation of char in the reactor bed. By the time the measurements were taken for this analysis, a significant amount of char had probably built up in the bed and was reacting with steam adding heating content to the syngas. To account for this in the energy flow, the difference was subtracted from the tar and char heat content value. This gave the most accurate reporting for steady-state gasification.

3.4 Gasification with Calcium Oxide

To further improve the energy conversion efficiency, a catalyst was used in the fluidized bed. This section presents the experimental findings of gasification performed with a CaO reactant. The temperatures tested were between 650 and 700 °C. At lower temperatures, low char reaction rates decrease biomass utilization and correspondingly the thermodynamic conversion efficiency. At higher temperatures, CO₂ reaction rates with limestone decrease. A full parametric study to determine the effect of S/B ratio or gas residence time was not performed. Test conditions were chosen based on the gasifier behavior observed in un-catalyzed testing. Moderate S/B ratios of between 2 and 3 were chosen to balance hydrogen production efficiency and energy efficiency. It was observed without a reagent, higher S/B ratios corresponded to increasing hydrogen production efficiency but lowered energy efficiency. The biomass feed rates were kept low to increase the duration of testing. Testing times were limited by the mass of limestone in the reactor. When the limestone becomes saturated with CO₂, the CO₂ concentration returns to levels seen in un-catalyzed testing. It was desired during this study to look at the behavior of gasification before this transition occurs, when the gasifier is at a quasi-steady state.

Once all of the CaO has reacted, it must be regenerated to continue testing. The ideal way to do this is to implement a second reactor. In this case, the spent quicklime is carried to the second reactor, regenerated, and returned to the fluidized gasification bed. With a single fluidized bed, as the case for this study, CaO gasification requires cycling the bed. This results in three distinct phases occurring as follows:

Biomass gasification phase: In this phase, biomass is inserted into the fluidized bed and gasified at temperatures between 650 and 700 °C. CO₂ produced during gasification reacts with CaO to produce CaCO₃, removing CO₂ from the reactor. The resulting shift in gas-phase composition creates more elemental hydrogen. At temperatures at and below 700 °C, the reactivity of char is low. As a result, char

builds up in the reactor during gasification. The volume of char increases until its total surface area is large enough to react at a rate equal to what is being produced by the addition of biomass feedstock.

Char gasification phase: In this phase, no biomass was inserted into the fluidized bed but steam still flowed through it. The steam slowly gasified the solid char, producing H_2 , CO , CO_2 , and CH_4 . The limestone reacted with CO_2 , provided that there was still unreacted CaO , and the CO_2 partial pressure was higher than the equilibrium pressure. At $650\text{ }^\circ\text{C}$, the equilibrium partial pressure of CO_2 is 0.8 % of atmospheric pressure and at $700\text{ }^\circ\text{C}$, it is 2.6 %.

CaO regeneration phase: In this phase, the temperature of the bed was increased to promote the release of CO_2 . The temperature required to raise the equilibrium pressure of CO_2 above atmospheric pressure is roughly $900\text{ }^\circ\text{C}$. However, if CO_2 is removed from the environment, lowering its partial pressure, CaO will regenerate at lower temperatures. The dry gas stream produced in this phase consists mainly of CO_2 and small amounts of hydrogen and carbon monoxide.

The resulting gas concentrations from gasification using CaO are shown in Table 4. The maximum average hydrogen concentration seen was 69.4 %. This was higher than any concentration seen without the use of CaO . The reason for the high H_2 levels was the elimination of CO_2 . During the maximum H_2 concentration test, the CO_2 concentration was 12.0 mol%, which is lower than any concentration seen during un-catalyzed testing. The CO concentration at the maximum H_2 concentration case was 7.5 %. This is nearly three times lower than the lowest levels seen during un-catalyzed testing. The low carbon monoxide levels were a result of Le Chatelier's principle working on the CO -shift reaction. This confirms that the CO -shift reaction is one of the most important reactions determining the syngas composition.

The effect of CaO on hydrocarbon levels varied. All the hydrocarbon concentrations measured during catalytic testing were similar to levels seen at $850\text{ }^\circ\text{C}$ in

Table 4 Results from gasification with CaO

Test #	1	2	3
Reactor temperature ($^\circ\text{C}$)	657	690	701
S/B ratio	2.9	2.9	2.1
Gas residence time (s)	2.7	2.6	2.7
Syngas component	Average gas conc. (dry mol%)		
Hydrogen	65.5	69.4	68.3
Methane	11.1	8.8	8.7
Carbon monoxide	10.8	7.5	9.3
Carbon dioxide	9.4	12.0	11.3
Ethylene	1.6	1.3	1.4
Ethane	0.5	0.8	0.6
Acetylene	0.1	0.1	0.1
Propylene	0.4	0.8	0.4
HHV (dry MJ/m^3)	15.6	14.2	14.3

un-catalyzed testing except for propylene which was higher than levels seen at 850 °C. The increased cracking of tars into propylene may have caused this. The gas filters were noted during testing to be cleaner than during un-catalyzed testing. This suggests that tar levels were the lowest when using CaO, although this cannot be proven because tars were not directly measured in this study. The fact that hydrocarbon levels were not lower than the concentrations seen at 850 °C also tells us that CO₂ and CO are not a product of hydrocarbon cracking. Other catalyst types are needed to break down hydrocarbons, which will form additional elemental hydrogen.

The results shown in Table 4 were averaged over the entire test. Figure 14 shows the gas concentration for over 2 h of testing. The hydrogen concentration decreased slightly, and the CO concentration increased over the course of the test. This was due to the saturation of the CaO. As the fraction of reacted CaO increased, the CO₂ removal rate decreased. If the gasifier was to be continued to be run, the hydrogen concentration would eventually drop to the levels seen during un-catalyzed testing.

The thermodynamic efficiencies of gasification are shown in Table 5. The highest cold gas efficiency seen was 76 %. This CGE was higher than any seen during un-catalyzed testing at 700 °C, but much less than the maximum values seen at 775 °C or 850 °C. The maximum energy efficiency seen was 47 %. This was the same maximum energy efficiency seen during un-catalyzed testing at 700 °C but much lower than the maximum values seen at higher temperatures. The low temperatures required for proper CaO reactivity prevented the full utilization of the biomass feedstock in this phase of operation.

A complete breakdown of the thermodynamic inputs/outputs for the maximum energy efficiency case is shown in Fig. 15. The biomass heat content accounted for 61 % of the input energy. Heating the superheated steam before it entered the fluidized bed consumed 23 % and external reactor heating 6 %. Unlike previous testing, these were not the only energy inputs. The reaction between CaO and CO₂

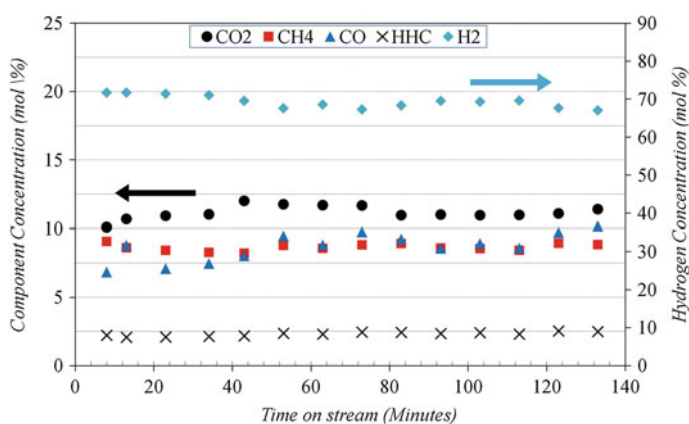
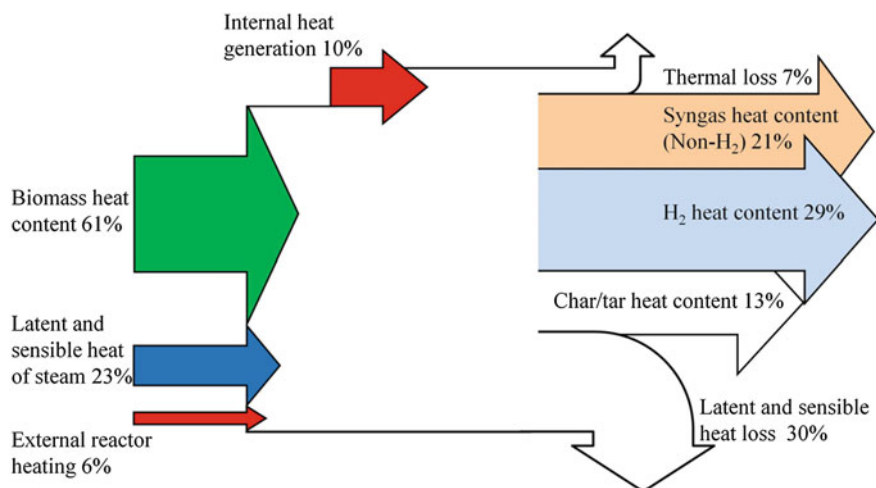


Fig. 14 Syngas composition for gasification using CaO (test number 3, $T = 701$ °C, $S/B = 2.1$, $\tau = 2.7$ s)

Table 5 Efficiency of gasification with CaO

Test #	1	2	3
Reactor temperature (°C)	657	690	701
S/B ratio	2.9	2.9	2.1
Gas residence time (s)	2.7	2.6	2.7
Cold gas efficiency (%)	56	70	76
Energy efficiency (%)	32	39	47
Hydrogen production efficiency (%)	16	23	27

**Fig. 15** Thermodynamic inputs/outputs for gasification with CaO for maximum efficiency case ($T = 701\text{ }^{\circ}\text{C}$, $S/B = 2.1$, $\tau = 2.7\text{ s}$)

releases 170.5 kJ/mol and added 10 % of the energy input to the system during this particular test. This heat content can be thought of as a battery, while energy was not inserted externally during testing, an equal amount of energy must be added during the regeneration phase. The energy addition from the CaO reaction was added to the energy efficiency and hydrogen production efficiency equations to give a more accurate measure of energy use during this phase. This is more useful when analyzing catalytic testing for use in a dual fluidized bed, which is its most likely use.

4 Summary and Conclusions

Currently, the most effective method for producing hydrogen-enriched synthesis gas from biomass is steam gasification. Steam gasification has been shown to produce a synthetic gas with high hydrogen concentration (>70 %) under certain

conditions, but little information exists on the thermal conversion efficiency during these conditions. Researchers have expressed doubt in steam gasification citing its large energy demand.

The syngas produced during fluidized bed steam gasification had a maximum hydrogen concentration of 51 mol%. During this test, the conversion of energy to syngas heating content and hydrogen heat content was 58 and 26 %, respectively. These values were obtained at the highest temperature tested (850 °C), a high steam-to-biomass ratio (2.9 by weight), and a high gas residence time in the reactor (4.1 s). A significant fraction of energy was needed to generate steam during this test, which limited the thermal conversion of energy to syngas. At the lowest steam-to-biomass ratio tested (0.75), the conversion of energy to syngas heat content was 68 % but the conversion to hydrogen was only 20 %. This was due to the relatively low hydrogen concentration (38 mol%).

Steam gasification with CaO produced the highest H₂ concentrations tested in this study (69 mol%). The maximum thermal conversion of energy to syngas heat content and hydrogen heat content was 47 and 27 %, respectively. The conversion efficiencies were lower than those observed during inert-bed testing because at the optimal temperature for CaO gasification (700 °C), and char conversion was low. During CaO regeneration, the char was gasified, producing a syngas stream with over 80 mol% hydrogen. The energy produced during regeneration equaled the energy requirement during this phase.

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